

FRACTIONAL DISTILLATION



APPLICATION OF THEORETICAL METHOD OF CALCULATION
TO COMMERCIAL GASOLINE STABILIZER
TEST DATA



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A Dissertation submitted in partial fulfillment of the requirements for the
degree of Doctor of Science in the University of Michigan,
Ann Arbor, Michigan.

1934



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III. APPLICATION OF THEORETICAL METHOD OF CALCULATION TO COMMERCIAL GASOLINE STABILIZER TEST DATA

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The greatest difficulty in attempting to check theoretical methods for the calculation of the number of plates required in a fractionating column, against actual operating data, is that of obtaining sufficiently complete and reliable data on the commercial plant. If simply the feed, overhead distillate, and bottoms are known as to quantity and quality, it is impossible to determine the actual operating efficiency of the plates in the column and therefore impossible to determine the equivalent number of equilibrium plates actually required to effect the separation.

Test Data

The test data used in this comparison include not only these essential data but also the complete analysis of the liquid samples as withdrawn from 17 of the 48 plates. Fig. 8 is a simplified flow diagram of the gasoline stabilizer, with the quantity of all significant streams indicated. Table VIII gives the composition of feed, bottoms, distillate, reflux, side-cut and liquid on various plates with the corresponding temperatures. Fig. 9 shows the temperature distribution in the column with all actual measurements indicated. The dotted line is an estimate of the temperature of the vapor at the top of the column based upon a single temperature reading of 130°F. for the vapors above the top tray. The important point is the effect of cold liquid reflux on the temperature of the liquid at the top of the column rather than the exact number of plates required for substantial equalization of the temperatures of the liquids and vapors.

In Fig. 10 the composition of the liquid plate samples is plotted in such a manner that the mole fraction of any one component is the difference between the values given between two points. This means

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that the mole fraction of any one component on any plate is represented by the difference between two lines rather than the value of a particular line. The areas formed between these lines are labelled

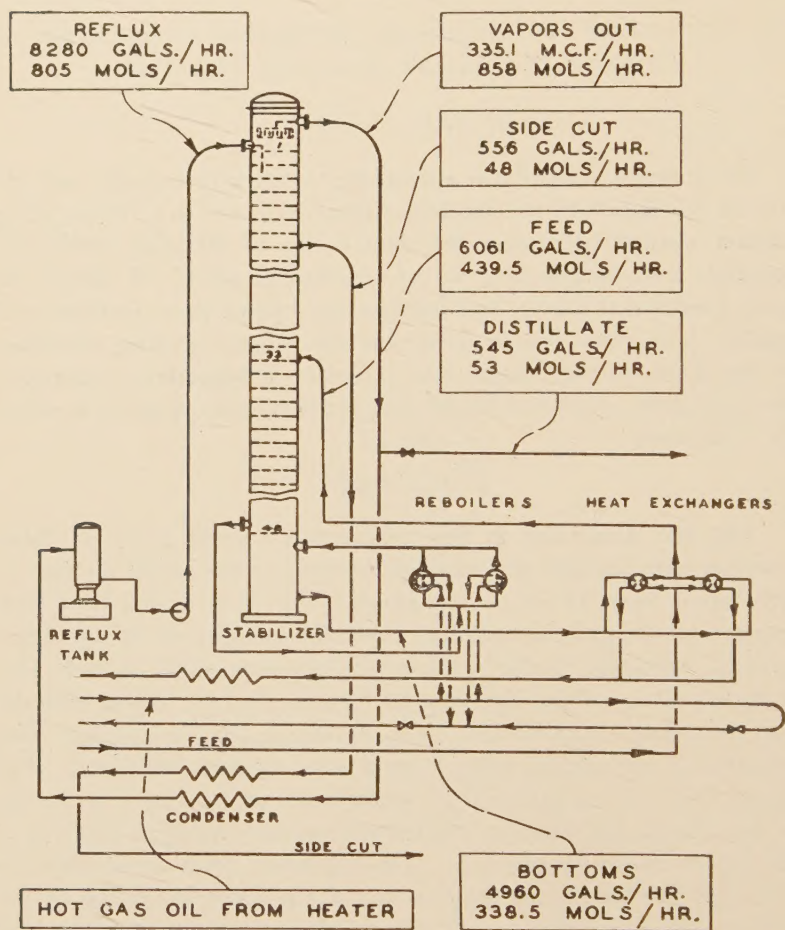


FIG. 8

with the name of the corresponding component. The lines drawn through the points represent the levelled data which has been used in the following computation. Except for the separation between iso and normal butane, and between iso and normal pentane in a few

samples from plates in the lower part of the column, the data appear unusually consistent and reliable.

In Fig. 11 the composition of the vapor rising from the plate is plotted in a similar manner, the circles indicating the vapor composi-

TABLE VIII. TEST DATA FOR GASOLINE STABILIZER

	Feed	Bottoms	Distil- late	Reflux	Distil- late plus side cut	Trays Below Feed				
						46	44	42	40	36
CH ₄ (mol fraction).....	0.0029		0.0258		0.0135					
C ₂ H ₆	0.0382		0.1335		0.0732					
C ₃ H ₈	0.1226		0.782		0.5875			0.0014	0.0052	.0239
iso-C ₄ H ₁₀	0.0798	0.0347	0.0527		0.2900	.0907	.0874	0.0966	0.1489	.1577
n-C ₄ H ₁₀	0.2752	0.3295	0.0035		0.0357	.5731	.6311	0.662	0.5867	.5452
iso-C ₅ H ₁₂	0.103	0.1303	0.0025(?)			.1051	.0944	0.0743	0.078	.0699
n-C ₅ H ₁₂	0.0929	0.1125				.0821	.0631	0.0277	0.0573	.0611
C ₇ H ₁₆	0.2854	0.393				.149	.124	0.138	0.1239	.1422
Reflux Ratio										
$\left(\frac{L}{D}\right)$										
$\left(\frac{L}{V}\right)$						1.32		1.325		1.33
Quantity (mols/hr.).....	439.5	338.5	53	805	101					
Temperature (°F.).....	255	330	130	72		278	272	268	266	262
Trays.....								15		

Trays Above Feed

	32	30	24	22	20	16	14	12	10	Side- cut 6	4	2
CH ₄ (mol fraction)											0.0180	0.0184
C ₂ H ₆	.0054	.009	.0022	.0012	.0029	.0032	.0021	.0016	0.003	0.0033	0.0243	0.1053
C ₃ H ₈	.0902	.0839	.0897	.0898	.0905	.0889	.1026	.0857	0.1204	0.3697	0.780	0.8085
iso-C ₄ H ₁₀	.195	.2514	.264	.3042	.3586	.4365	.4794	.5408	0.5896	0.5524	0.1774	0.0638
n-C ₄ H ₁₀	.5744	.6192	.6403	.6036	.5449	.4694	.4124	.3688	0.287	0.0713	0.0165	0.004
iso-C ₅ H ₁₂	.0536	.0214	.0038	.0012	.0031	.002	.0035	.0031		0.0053		
n-C ₅ H ₁₂	.0375	.0106										
C ₇ H ₁₆	.0439	.0045										
Reflux Ratio												
$\left(\frac{L}{D}\right)$	8.06	9.35	9.45	9.45	9.45	.945	9.45	9.45	9.45		19.2	16.6
$\left(\frac{L}{V}\right)$	.89	.903	.905	.905	.905	.905	9.05	.905	0.905		0.95	0.943
Quantity (mols/hr.).....	235	225	221	220	219	217	216	213	209	184	161	116
Temperature (°F.).....						32						
Trays.....												

tion as computed from actual liquid analysis by means of equilibrium relationships based upon ideal solutions, and the triangular points represent the composition of the vapor as computed from heat and material balances around parts of the column. The lines represent

the levelled data as in the case of Fig. 10. Comparison of the vapor composition as computed by the two different methods indicates that

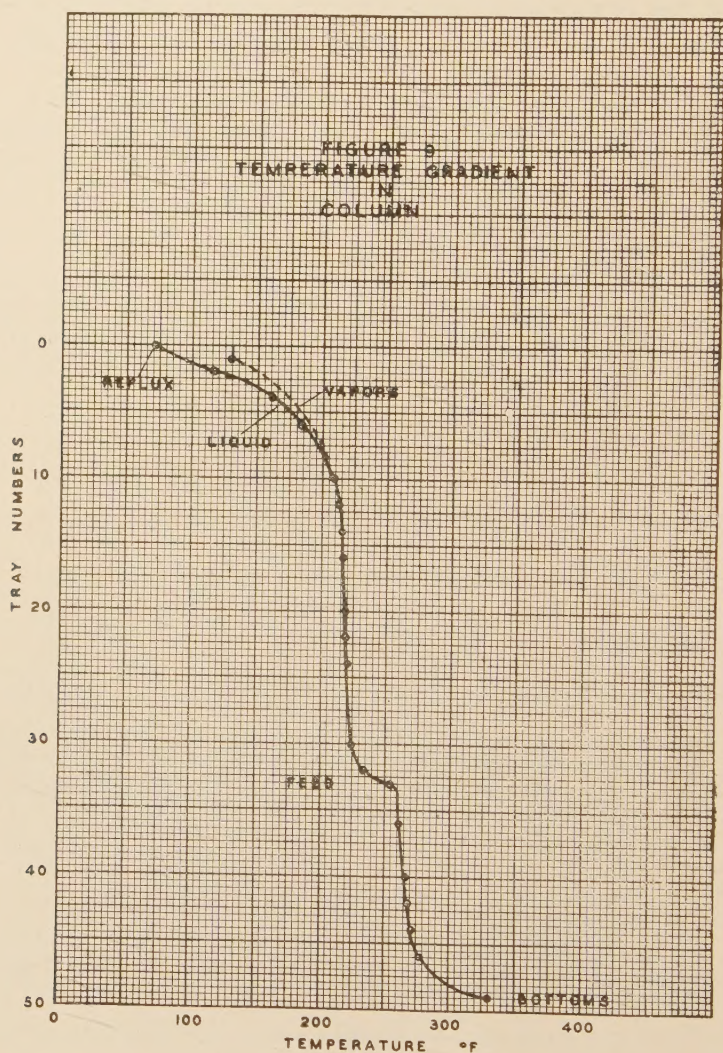


FIG. 9

the composition of the vapor as determined by heat and material balances is substantially the same as that computed from equilibrium data and the composition of the liquid. Therefore the plate efficiency

in this column under these conditions of operation is very close to 100%. In other words each actual plate is just about as effective as an equilibrium plate.

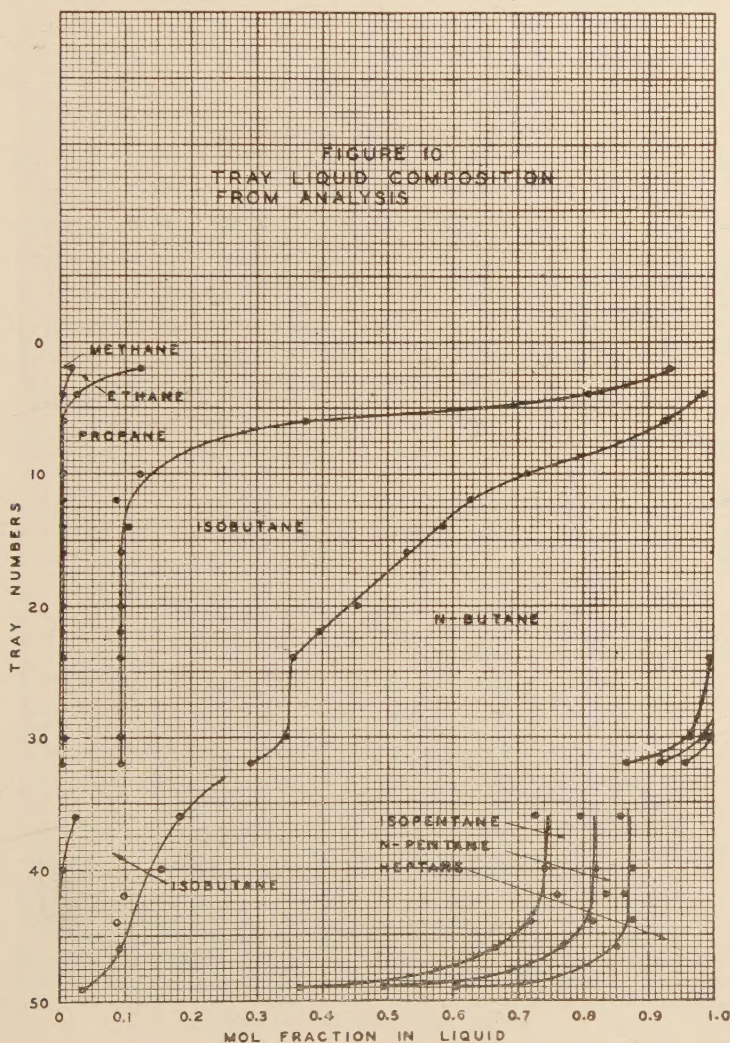


FIG. 10

In Fig. 12 the liquid overflow (L) in moles per hour is plotted for the various plates in the column. The ratio of liquid overflow to

vapor rising in the column (L/V) as calculated for various parts of the column is given in Table VIII and is practically constant in the

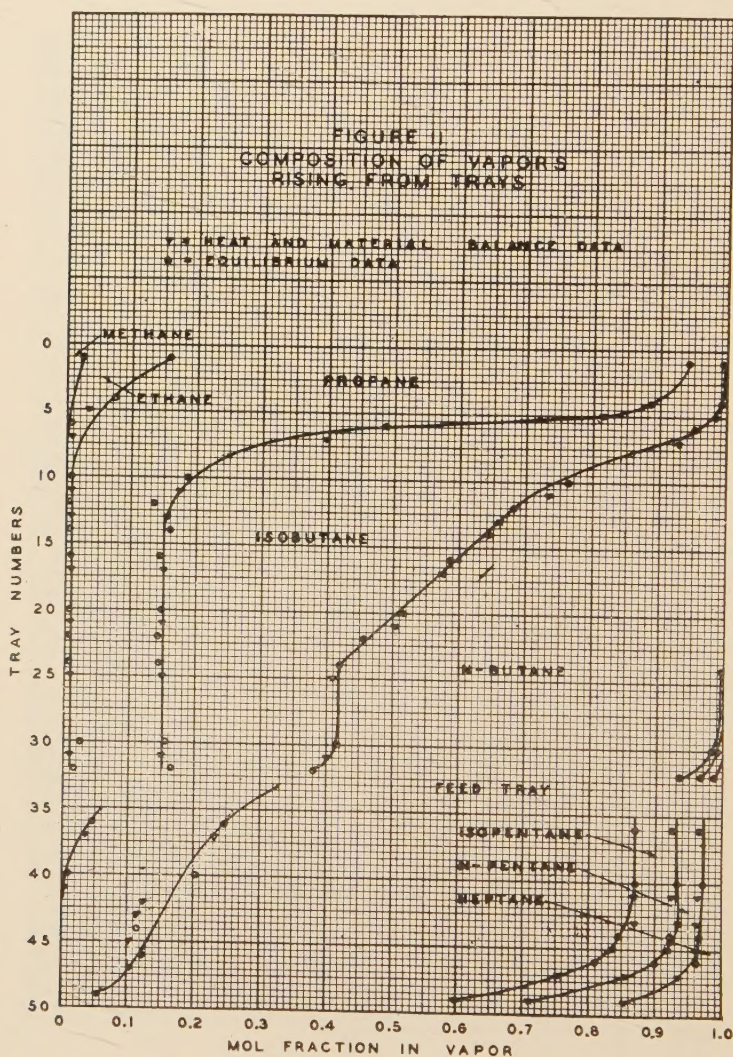


FIG. 11

stripping section (plate 34 to 48) and in the rectifying section below the side-stream (plate 7 through 32), with the exception of the

difference immediately above and below the feed plate. Above the side-cut the disturbing effects of the cold reflux is clearly evident.

Actual Individual Plate Efficiency

The most generally accepted definition of plate efficiency is that given by Murphree⁵² and represented by the equation,

$$E_y = \frac{y_n - y_{n+1}}{y_n^* - y_{n+1}} = \frac{y_n - y_{n+1}}{K_n x_n - y_{n+1}} \quad (25)$$

where

y_n = mol fraction of a component in vapors rising from a plate.

y_{n+1} = mol fraction of the component in vapors rising to the plate.

y_n^* = mol fraction of the component in vapors in equilibrium with the liquid leaving the plate = $K_n x_n$.

Similarly, the same plate efficiency defined in terms of the change in liquid compositions is represented by the equation,

$$E_x = \frac{x_{n-1} - x_n}{K_n \frac{V}{L} x_n - \frac{D}{L} x_D - x_n} \text{ or } \frac{x_{m-1} - x_m}{K_m \frac{\bar{V}}{L} x_m + \frac{B}{L} x_B - x_m} \quad (26)$$

Other definitions have been used and frequently erroneously assumed to be identical with Equation (25) or (26), such as,

$$\frac{x_{n-1} - x_n}{x_{n-1} - x_n^*} = \frac{x_{n-1} - x_n}{x_{n-1} - \frac{y_n}{K_n}}$$

where

x_n = mol fraction of a component in liquid leaving a plate.

x_{n-1} = mol fraction of the component in liquid entering the plate.

x_n^* = mol fraction of the component in liquid in equilibrium with the vapors rising from the plate.

TABLE IX. INDIVIDUAL PLATE EFFICIENCIES (MURPHREE'S)

	Plate 12 (214° F.)				Plate 15 (216° F.)		Plate 20 (218° F.)		Plate 32 (233° F.)	
	C ₃ H ₈	iC ₄ H ₁₀	nC ₄ H ₁₀	nC ₄ H ₁₀	iC ₄ H ₁₀	nC ₄ H ₁₀	iC ₄ H ₁₀	nC ₄ H ₁₀	iC ₄ H ₁₀	nC ₄ H ₁₀
$L/V = .905$										
$L/D = 9.45$										
x_{n-1}	.12	.545	.335	.483	.420	.370	.537	.220	.605	
x_B	.107	.521	.372	.462	.444	.347	.560	.197	.573	
x_D	.5875	.290	.0357							
$D/L x_D$	.0622	.0307	.00378							
$D/L x_D + x_B$	.169	.552	.376	.493	.448	.378	.564	.228	.577	
$x_{n-1} - x_B$	+.013	+.024	-.037	+.024	-.024	+.023	-.023	+.023	+.032	

Equilibrium Constants (K) ⁶⁷

K_{n-1}	1.5	.99	.85	1.00	.865	1.02	.88	1.11	.95
$K_{nV/L} x_B$	.177	.57	.349	.511	.425	.392	.545	.242	.601
Equib. Plate									
$x_{n-1} - x_B$	+.008	+.018	-.027	+.018	-.023	+.014	-.019	+.014	+.024
Ex.	1.62	1.33	1.37	1.33	1.04	1.64	1.21	1.64	1.33

K corrected for Deviation from Ideal Solution ³²

A.	1.14	1.047	1.068	1.056	1.057	1.073	1.042	1.105	1.04
$K_{nV/L} x_B$	.202	.595	.372	.54	.449	.42	.568	.267	.624
Equib. Plate									
$x_{n-1} - x_B$	+.033	.043	-.004	+.047	+.001	.042	+.004	.039	+.047
Ex.	.394	.557	9.	.51	-24.00	.55	-5.75	.59	.68

Any effort at defining individual plate efficiency must be arbitrary, but usage and logic are both in support of E_y or E_x as defined in Equations (25) and (26). In commercial work the overall plate efficiency, defined as the ratio of the number of equilibrium plates to number of actual plates required for the same separation is of more practical value and can be determined with less error. As shown by Holbrook and Baker the overall plate efficiency may be more or less than the Murphree or individual plate efficiency, depending upon the nature of the equilibrium curve and operating line. The overall and individual plate efficiencies are equal only when the equilibrium curve and operating line are straight and parallel, or when the individual plate efficiency is unity.

For these reasons it is suggested that only two definitions of plate efficiency be accepted, Murphree's⁵² individual plate efficiency as defined by Equations (25) and (26) and the overall plate efficiency defined as the number of equilibrium plates divided by the number of actual plates.

Individual plate efficiencies were calculated by Equation (26) from actual data given in Table VIII and interpolations thereof from Fig. 10. The individual plate efficiencies calculated in this manner are given in Table IX.

Plate efficiencies were calculated on the basis of the equilibrium constants⁷¹ which have been used in all of the calculations in this paper, and also on the basis of the same equilibrium constants as corrected for deviations from ideal solutions according to the estimated correction factor of Katz and Brown.³³ The computation of the individual plate efficiencies by Equation (26) is a rather unsatisfactory procedure. Due to the fact that the efficiency is determined by subtracting two numbers of approximately the same order of magnitude, the method is extremely sensitive to errors in the data and errors in the equilibrium relationship.

Although the individual plate efficiency is of considerable theoretical significance, in the practical design and operation of commercial columns the overall or average plate efficiency is of much greater significance.

The individual plate efficiency as indicated in Table IX averages somewhat over 100%, and is relatively more consistent on plates above the feed plate and when the uncorrected equilibrium constants are used. Substantially the same results are indicated in Fig. 11 in

which the vapor in equilibrium with the liquid was computed by use of the uncorrected equilibrium constants.

As suggested by Katz and Brown,³³ the use of the corrections with the equilibrium constants is not justified in the computation of plate efficiency and the design of fractionating equipment, at least until more information is available concerning the deviation from ideal solutions. On the other hand the high plate efficiencies indicated in some cases when no correction factor is used, are without doubt fictitious as they are invariably greatly reduced when a correction factor is applied to the equilibrium constant.

*Algebraic Method*⁴⁴

This column was analyzed, first of all, by attempting to work down from the given distillate composition and up from the given liquid composition on plate 46. Plate 46 was chosen as the starting point in preference to the bottoms because the liquid bottoms are not in equilibrium with the vapors rising into plate 48. The heating requirements are furnished by recirculating liquid from plate 48 through a heat exchanger, the vapors entering the column just below the plate. This arrangement upsets the equilibrium between liquid bottoms and vapors entering the bottom plate.

As an example of the calculations, a temperature is assumed for the liquid on plate 46, and the composition of the vapors in equilibrium with this liquid computed from the equilibrium constant (K) values at the assumed temperature and the pressure of the column. If Σy is unity the correct temperature has been assumed.

N-hexane and heavier components are considered as the single component, n-heptane.

	x_{46} mol fraction	K 283° F. (300 lbs. per sq. in. abs.)	y_{46} mol fraction
iso-C ₄ H ₁₀	0.0907	1.34	0.1216
n-C ₄ H ₁₀	0.5731	1.20	0.6875
iso-C ₅ H ₁₂	0.1051	0.865	0.0910
n-C ₅ H ₁₂	0.0821	0.76	0.0624
C ₇ H ₁₆	0.1490	0.265	0.0394
	1.0000		1.0019

At a temperature of 283° F. the mol fractions in the vapors rising from plate 46 into plate 45 are now known. The composition of the

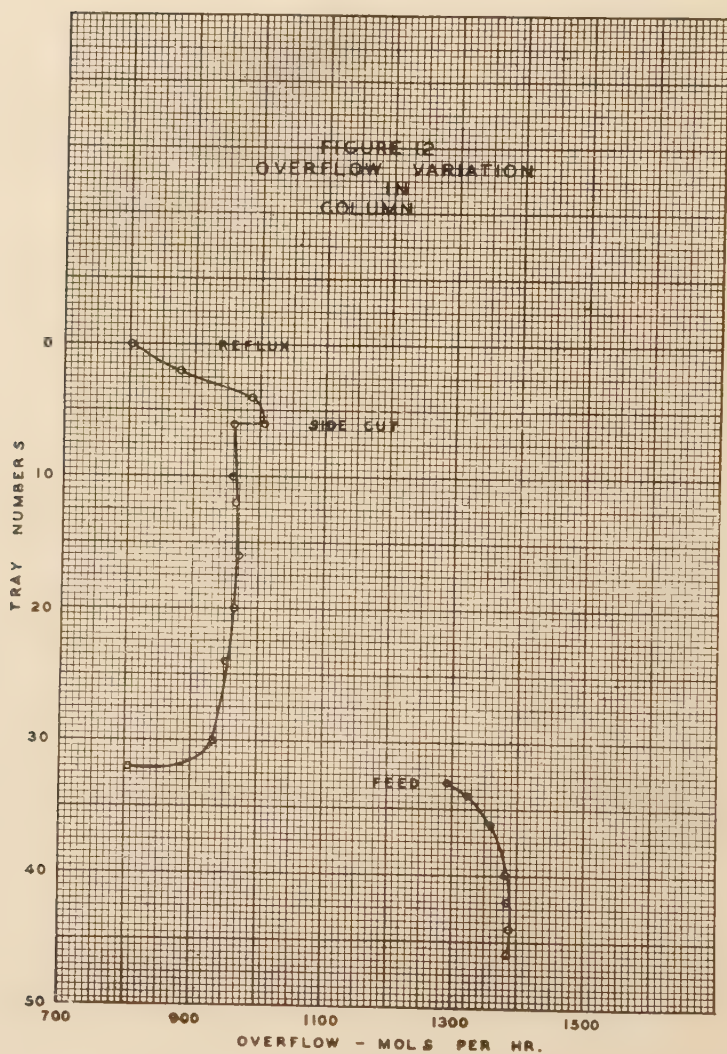


FIG. 12

liquid on plate 45 can then be calculated from a material balance using Equation (9):

$$x_{m-1} = \frac{\bar{V}}{\bar{L}} y_m + \frac{B}{\bar{L}} x_B$$

for

$$\text{iso-C}_4\text{H}_{10}, x_{45} = 0.754 (0.1216) + 0.246 (0.0347) = 0.100$$

$$\text{n-C}_4\text{H}_{10}, x_{45} = 0.754 (0.6875) + 0.246 (0.3295) = 0.5988$$

$$\text{iso-C}_5\text{H}_{12}, x_{45} = 0.754 (0.091) + 0.246 (0.1303) = 0.1005$$

$$\text{n-C}_5\text{H}_{12}, x_{45} = 0.754 (0.0624) + 0.246 (0.1125) = 0.0746$$

$$\text{C}_7\text{H}_{16}, x_{45} = 0.754 (0.0394) + 0.246 (0.393) = 0.1261$$

1.000

No methane, ethane, or propane appears in the analysis of the liquid on plate 46. Therefore, some value must be assumed for these components in order that the calculations may be carried on to the feed plate with some degree of accuracy. A small amount of propane (0.14 per cent) was arbitrarily inserted in the liquid on plate 42 and the computations continued on up to the feed plate. Methane and ethane have a much smaller mol fraction in this section of the column and have been disregarded, although they also become of some importance in the vicinity of the feed plate.

Table X gives the results of the calculations which are plotted in Figures 13 and 14 with tray number as ordinates and mol fraction in the liquid as abscissae. The small triangles represent calculated values while the circles are experimental points.

The calculated curves agree fairly well with the experimental values, especially for the heavier components, indicating a plate efficiency of approximately 100%.

There is uncertainty, however, in the calculations of the upper part of the stripping column due to the more volatile components, methane, ethane and propane, the composition of which must be assumed.

Similar calculations were made for the rectifying section of the column, starting with the composition of vapors leaving the top of the column. Although 0.25 per cent of isopentane was reported in the distillate, there can be no measurable quantity of isopentane, pentane, or hexane present because none was found on plates 2 or 4. If 0.25 per cent of isopentane were present in the distillate, all of the

analyses of the upper plates would be in error. In the calculations no pentane was considered present until tray 21 was reached, at which small amounts of isopentane and n-pentane (0.02 per cent and 0.01 per cent respectively) were arbitrarily added to the liquid composition on that plate. The results of this effort are shown in Fig. 14 by the small triangles. For the upper part of the rectifying section where the less volatile components are absent and the uncertainty concerning them is removed, the calculated and experimental curves agree well for all components.

In the lower part of the rectifying column, where the less volatile components appear, the curves deviate due to the uncertainty in estimating the mol fractions of the less volatile constituents. The deviation of the computed from the observed compositions is even more pronounced in the dotted curves (Fig. 13) for isobutane and n-butane, which were calculated by introducing a greater quantity of isopentane (0.2 per cent at tray 15) and a greater quantity of n-pentane (0.1 per cent at tray 17). Heptane has been completely ignored, but it also would become of importance in the calculations near the feed plate.

When working from the top and bottom compositions toward the feed, no method is available by which the components less volatile than the heavy cut-component may be correctly incorporated into the computations of the rectifying section, and no means is known by which those components more volatile than the lighter cut-component may be correctly added to the calculations of the stripping section.

Efforts to estimate the composition on the feed plate by use of Equations (13) and (14) as has been suggested⁴⁴ were of little or no help in obtaining a satisfactory solution to the column calculations. Although practically the only method available for the complete calculation of columns from a theoretical standpoint, this method is not satisfactory for practical engineering work because of the large amount of calculation required and because of the relatively indefinite results obtained.

The Absorption Factor Method

In the second part of this study it was shown that the absorption factor method could be applied to the calculation of columns with substantially the same results as obtained from the more laborious plate-to-plate calculations. The following calculation gives the results obtained from applying the absorption factor method of calculation to an actual commercial gasoline stabilizer.

TABLE X. RESULTS OF ALGEBRAIC METHOD
MOL FRACTION IN LIQUID OR VAPOR

Temp. (°F.)	y ₁	x ₁	y ₂	x ₂	y ₃	x ₃	y ₄	x ₄	y ₅	x ₅	y ₆	x ₆	y ₇	x ₇	y ₈	x ₈	y ₉	x ₉
	133		146		155		163		172		182		188		197		202	
CH ₄	0.0258	0.0021	0.00345	0.00026	0.0017	0.00012	0.00158	0.0001	0.00156	0.0000945	0.00154	0.00087	0.0013	0.000067	0.00128	0.000064	0.0127	0.000058
C ₂ H ₆	0.1335	0.069	0.0726	0.0356	0.0412	0.0195	0.026	0.0119	0.0188	0.00835	0.0755	0.0066	0.0125	0.00515	0.0111	0.0044	0.105	0.0041
C ₃ H ₈	0.782	0.822	0.8196	0.788	0.7876	0.716	0.7196	0.623	0.6326	0.518	0.5331	0.4185	0.434	0.324	0.348	0.252	0.283	0.20
iso-C ₄ H ₁₀	0.0527	0.099	0.0964	0.1593	0.1533	0.232	0.2219	0.311	0.296	0.39	0.371	0.4555	0.4412	0.515	0.4952	0.55	0.572	0.569
N-C ₄ H ₁₀	0.0035	0.0085	0.0082	0.0175	0.0167	0.0321	0.0305	0.054	0.0511	0.0837	0.0792	0.119	0.1114	0.156	0.1446	0.1926	0.1779	0.227
iso-C ₅ H ₁₂																		
N-C ₅ H ₁₂																		
C ₇ H ₁₆																		
Total.....	0.9975	1.000	1.000	1.000	1.000	0.999	0.999	1.000	1.000	1.000	1.000	0.999	1.000	1.000	1.000	0.999	0.999	1.000

MOL FRACTION IN LIQUID OR VAPOR

Temp. (°F.)	y ₁₀	x ₁₀	y ₁₁	x ₁₁	y ₁₂	x ₁₂	y ₁₃	x ₁₃	y ₁₄	x ₁₄	y ₁₅	x ₁₅	y ₁₆	x ₁₆	y ₁₇	x ₁₇	y ₁₈	x ₁₈
	207		209		211		213		214		215		216		216.5		217	
CH ₄	0.00126	0.000057	0.00126	0.000055	0.0099	0.0037	0.00986	0.0037	0.00983	0.00365	0.0098	0.00362	0.00976	0.00361	0.00975	0.0036	0.00974	0.00359
C ₂ H ₆	0.0102	0.0039	0.01	0.0038	0.1807	0.1224	0.166	0.1118	0.1561	0.1045	0.1495	0.0996	0.1452	0.0965	0.1424	0.0945	0.1405	0.093
C ₃ H ₈	0.236	0.1633	0.2028	0.1388	0.5432	0.556	0.5322	0.541	0.5192	0.525	0.5022	0.503	0.4847	0.483	0.4667	0.4635	0.4487	0.4445
iso-C ₄ H ₁₀	0.5442	0.574	0.5487	0.568	0.266	0.318	0.2916	0.344	0.3149	0.369	0.3376	0.393	0.3596	0.416	0.3801	0.4385	0.4006	0.46
N-C ₄ H ₁₀	0.2091	0.26	0.239	0.29														
iso-C ₅ H ₁₂																		
N-C ₅ H ₁₂																		
C ₇ H ₁₆																		
Total.....	1.000	1.001	1.001	1.000	1.001	1.000	1.000	1.000	1.001	1.000	0.999	0.999	0.999	0.999	0.999	1.000	0.999	1.001

TABLE X. RESULTS OF ALGEBRAIC METHOD
MOL FRACTION IN LIQUID OR VAPOR

Temp. (°F.)	y ₁₉	y ₂₀	y ₂₁	x ₂₁	y ₂₂	x ₂₂	y ₂₃	x ₂₃	y ₂₄	x ₂₄	y ₂₅	x ₂₅	y ₂₆	x ₂₆	y ₂₇	x ₂₇
218		219		219.5	220		220.5		221		222		222.5		223	
CH ₄ ...	0.00973	0.00457	0.00971	0.00454	0.00968	0.00452	0.00967	0.00451	0.00966	0.0045	0.00965	0.0045	0.00965	0.00448	0.00964	0.00446
C ₂ H ₆ ...	0.192	0.0919	0.182	0.0905	0.171	0.0896	0.156	0.0885	0.131	0.088	0.1347	0.0874	0.1341	0.0868	0.1336	0.0861
iso-C ₃ H ₈ ...	0.437	0.475	0.4147	0.407	0.4012	0.399	0.382	0.3715	0.354	0.3497	0.338	0.322	0.307	0.302	0.292	0.292
N-C ₃ H ₈ ...	0.4199	0.48	0.4381	0.499	0.4716	0.535	0.681	0.552	0.5036	0.5675	0.5176	0.5815	0.501	0.592	0.596	0.60
iso-C ₄ H ₁₀ ...					0.0036	0.00068	0.0061	0.011	0.01	0.0193	0.01175	0.01525	0.02293	0.00538	0.0188	0.00887
N-C ₄ H ₁₀ ...					0.0018	0.00039	0.0035	0.0076	0.0089	0.0147	0.00133	0.01282	0.02355	0.00737	0.0186	0.01
Total.....	1.000	1.000	1.000	1.000	0.999	1.000	0.999	1.000	0.999	1.000	1.000	1.000	0.999	1.000	0.999	1.000

MOL FRACTION IN LIQUID OR VAPOR

Temp. (°F.)	y ₂₈	x ₂₈	y ₂₉	x ₂₉	y ₃₀	x ₃₀	y ₃₁	x ₃₁	y ₃₂	x ₃₂	y ₃₃	x ₃₃	y ₃₄	x ₃₄	y ₃₅	x ₃₅
225		227.5		230		265		266.5		267		268		269		270.5
CH ₄ ...	0.00961	0.00343	0.00958	0.0034	0.00956	0.00335	0.00952	0.0033	0.00947	0.00326	0.00942	0.00317	0.00932	0.00308	0.00927	0.00304
C ₂ H ₆ ...	1.829	0.852	1.721	0.839	1.61	0.8219	0.197	1.564	0.196	1.54	0.193	1.515	0.19	1.485	0.1856	1.44
iso-C ₃ H ₈ ...	2.932	2.76	2.792	0.267	2.662	0.243	0.62	2.588	0.614	2.488	0.608	2.385	0.601	2.288	0.5948	2.197
N-C ₃ H ₈ ...	0.466	0.02	0.486	0.095	0.526	0.076	0.62	0.576	0.61	0.757	0.076	0.881	0.076	0.981	0.076	1.081
iso-C ₄ H ₁₀ ...	0.08	0.143	0.13	0.0246	0.0205	0.015	0.0397	0.015	0.0382	0.0171	0.0392	0.0382	0.0382	0.0382	0.0382	0.0382
N-C ₄ H ₁₀ ...	0.0014	0.128	0.17	0.14	0.108	0.061	0.0249	0.117	0.236	0.118	0.238	0.1161	0.0267	0.1161	0.0267	0.1161
Total.....	999	999	999	1.000	1.000	1.000	1.000	998	998	998	998	999	1.001	998	998	998

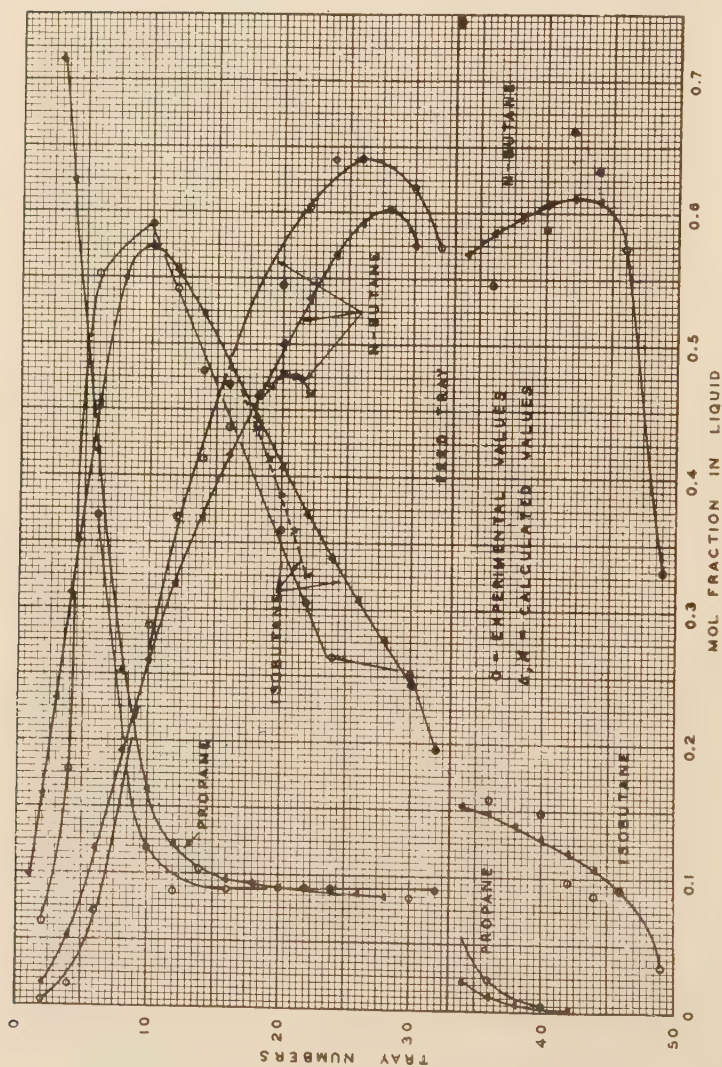


FIG. 13

Composition of Feed Plate

The composition of the liquid on the feed plate was calculated by applying Equation (13) to those components more volatile than isobutane and by applying Equation (14) to those components less volatile than normal butane. The quantity of distillate (D) is the sum of the overhead product and the sidestream, and the composition of the distillate is the composite of these two products. On this basis the compositions of the liquid and vapor at the feed plate were determined to be as follows:

COMPOSITIONS AT FEED PLATE—TEMPERATURE 256° F.

	x_f	K 300 lbs./sq. in. abs. 256° F. .	y_f
CH ₄	.00015	9.6	.00142
C ₂ H ₆	.00305	• 3.18	.00979
C ₃ H ₈	.0649	1.78	.1157
iso-C ₅ H ₁₂	.0698	.717	.0501
n-C ₅ H ₁₂	.0529	.63	.03235
C ₇ H ₁₆	.1129	.193	.0218
	0.3037		0.2312

Denoting iso-butane by the subscript d and normal butane by the subscript e

$$x_{df} + x_{ef} = 0.6963$$

$$1.205 x_{df} + 1.06 x_{ef} = 0.7688$$

$$x_{df} = 0.207$$

$$x_{ef} = 0.489$$

$$y_{ef} = 0.519$$

Because of the withdrawal of a side-stream and the effect of vaporization in the reboiler, the column should be divided into four sections; (1) the top six plates above the side-stream, (2) the section between the feed plate and the side-stream, (3) the section between the feed and lowest plate for which analytical data are available, and (4) the remainder including the reboiler.

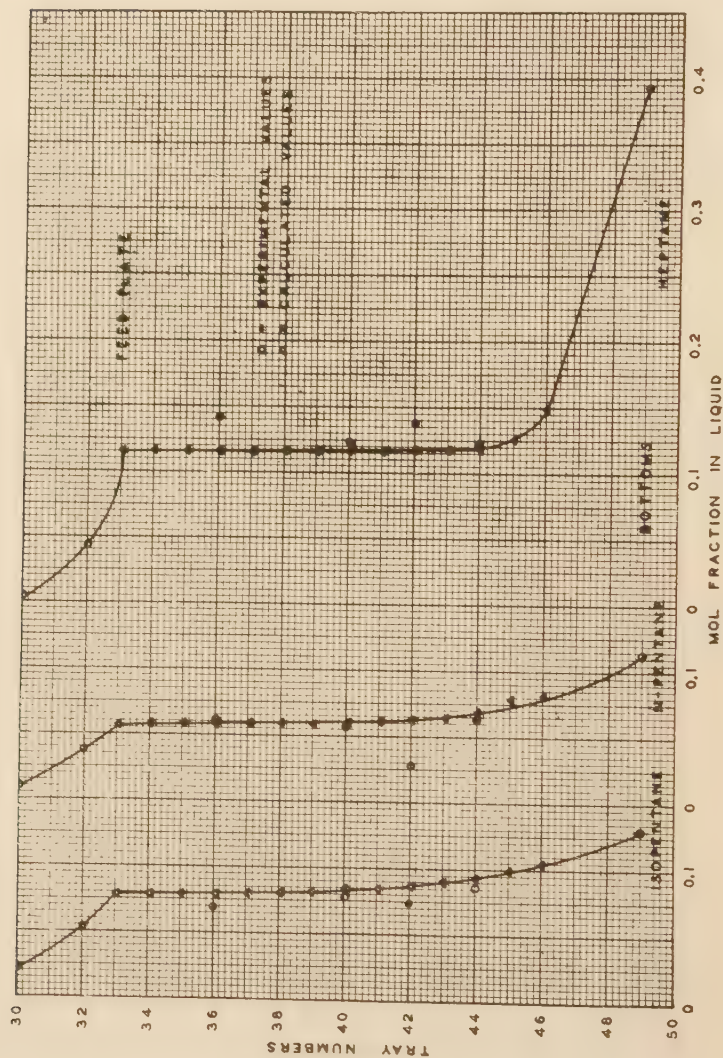


FIG. 14

Number of Plates Between Plate 7 and Top of Column

This section is operating as an absorber of n-butane. The mol fraction of n-butane in the vapor entering this section is obtained from a material balance around the top of the column and above plate 7;

$$y_{e7} = 0.905 \times 0.0713 + 0.0956 \times 0.0357 = 0.0649$$

The temperature of plate 6 was observed as 190° F. and the temperature of the top plate 130° F.

Applying the absorption factor method to n-butane above the side stream,

$$K_A = \frac{0.72 + 0.40}{2} = 0.55$$

$$A = \frac{L}{K_A V} = \frac{0.95}{0.55} = 1.727$$

$$\frac{y_7 - y_1}{y_7 - K_1 x_R} = \frac{y_7 - y_1}{y_7 - K_1 y_1} = \frac{0.0649 - 0.0035}{0.0649 - 0.4 \times 0.0035} = 0.967$$

Using this value in Equation (20) (Fig. 7) indicates that 5 equilibrium plates are required above the side-stream. Although there are six plates in this section of the column, two of them were dry during this test as the reflux was introduced on the third plate as indicated in Figure 8, and the number of actual working plates is uncertain. If it is assumed that the two dry plates did no work, the indicated overall plate efficiency is apparently 125%. But if it is assumed that the two dry plates were roughly equivalent to one normal plate, the overall plate efficiency is apparently 100 per cent.

Number of Plates Between Feed Plate and Plate 6

This section is operating as an absorber of n-butane. Assuming a linear temperature distribution the temperature of the plate above the feed plate is calculated as 254°F. The equilibrium temperature of plate 7 is 193°F.

Applying the absorption factor method to n-butane,

$$K_A = \frac{1.05 + 0.73}{2} = 0.89$$

$$A = \frac{L}{K_A V} = \frac{0.905}{0.89} = 1.017$$

$$\frac{y_f - y_7}{y_f - K_7 x_6} = \frac{0.5190 - 0.0649}{0.5190 - 0.0520} = 0.9724$$

Using this value in Equation (20) indicates that 28 equilibrium plates are required. Since there are 26 actual plates in this section, the overall plate efficiency is 108% on the basis of the values of the equilibrium constants used without any correction for deviation from ideal solutions.

Because the feed heater is not an equilibrium apparatus and because the actual feed plate is not an equilibrium plate, the vapor rising to the plate above is actually more volatile than the equilibrium vapor calculated for use with the absorption factor method. This results in actual temperatures much lower than the calculated temperatures for plates immediately above the feed, and a corresponding adjustment of the compositions so that the concentration of n-butane in the actual column passes through a maximum above the feed plate. This maximum is not introduced in calculations by the absorption factor method, and hence the absorption factor may be applied to n-butane in this part of the column if calculated temperatures rather than actual temperatures are used. This procedure should always be followed as in computing the number of equilibrium plates, equilibrium temperatures are of significance, rather than actual temperatures.

Number of Plates Between Feed Plate and Plate 47

This section is operating as a stripper of isobutane. The mol fraction of isobutane in the vapor entering this section (y_{d47}) is obtained from a material balance around the bottom product and between plates 46 and 47;

$$y_{d47} = 1.325 \times 0.0907 - 0.325 \times 0.0347 = 0.1089$$

The equilibrium temperature of plate 46 is calculated as 283°F., and the temperature of the plate below the feed plate as 258°F.

Applying the stripping factor method to isobutane,

$$K_s = \frac{1.34 + 1.23}{2} = 1.285$$

$$S = \frac{K_s \bar{V}}{\bar{L}} = \frac{1.285}{1.325} = 0.970$$

$$\frac{x_f - x_{46}}{x_f - \frac{y_{47}}{K_{46}}} = \frac{0.207 - 0.0907}{0.207 - \frac{0.1089}{1.34}} = 0.924$$

Using this value in Equation (20) indicates that 14 theoretical plates are required. With 13 actual plates, the overall plate efficiency is 108% on the basis of the ideal equilibrium constants.

Checking the assumed temperature of the feed plate (256°F.) by the linear temperature equation

$$t_f = 193 + \frac{28}{42} (283 - 193) = 193 + 60 = 253^\circ \text{ F.}$$

Number of Plates Below Plate 46

This section includes the reboiler and operates as a stripper of isobutane. The calculated equilibrium temperature at the reboiler is 350°F. and at plate 47 is 300°F.

Applying the stripping factor to isobutane,

$$K_s = \frac{1.67 + 1.42}{2} = 1.545$$

$$S = \frac{K \bar{V}}{\bar{L}} = \frac{1.545}{1.325} = 1.166$$

$$\frac{x_{46} - x_m}{x_{46} - \frac{y_{m+1}}{K_m}} = \frac{x_{46} - x_B}{x_{46} - \frac{x_B}{K_B}} = \frac{0.0907 - 0.0347}{0.0907 - \frac{0.0347}{1.67}} = 0.801$$

Using this value in Equation (22) indicates that 3 theoretical plates are required. There are 3 actual plates including the reboiler as one plate.

Correct Location of Feed Plate

Since this column is separating a complex mixture between two adjacent components, isobutane and normal butane, the correct location of the feed plate may be determined from Equations (18) and (18a). According to Equation (18a) the feed plate is correctly located when

$$\frac{x_{df}}{x_{ef}} > .4238$$

The value for this ratio on the feed plate is 0.4239. Similarly the terms of Equation (18) reduce to substantial equality with this feed plate composition. Since the inequalities become equalities for the special case of minimum reflux, we may conclude that the solution obtained by means of the absorption factor method is not only that solution which correctly indicates the operation of the column, but in this case is also the solution which requires the least number of plates for a given separation under the existing operating conditions. Because of the large number of plates included in the column, it is to be expected that the composition on the feed plate would approximate that obtained under conditions of minimum reflux (infinite number of plates).

Conclusions

From the comparison of actual plant test data with the computed results obtained by the application of the absorption factor method to the calculation of this gasoline stabilizer, it seems that the absorption factor method offers a simple and reliable method for computing the number of equilibrium plates required in commercial gasoline stabilizers, and probably other similar equipment in which the equilibrium conditions between liquid and vapor may be assumed to follow the relationships of ideal solutions.

NOMENCLATURE

- F = Moles of feed to column; used as subscript to signify feed.
B = Moles of bottom product; used as subscript to signify bottom product.
D = Moles of distillate withdrawn as overhead product; used as subscript to signify overhead product.

- V = Moles of vapor rising from plate in rectifying section; used as subscript to signify vapor phase.
 L = Moles of liquid overflowing from plate in rectifying section; used as subscript to signify liquid phase.
 $\bar{V}$ = Moles of vapor rising from plate in stripping section.
 $\bar{L}$ = Moles of liquid overflowing from plate in stripping section.
 a, b, c, d, etc., refer to moles of individual components of which a is the most volatile; when used as subscript the letters refer to individual components.
 x = Mole fraction of any component in the liquid.
 y = Mole fraction of any component in the vapor.
 n = Number of theoretical plates above feed plate; used as subscript, refers to any plate in rectifying section.
 m = Number of theoretical plates below feed plate; used as subscript, refers to any plate in stripping section.
 f = feed plate, i.e., plate from which the vapor rising becomes the feed to the rectifying section and liquid overflowing becomes the feed to the stripping section.
 M = Bottom plate of column.
 1, 2, 3, etc., refer to the first, second, third, etc., plate of the column, numbering from the top downward.
 P = Total pressure.
 p = Vapor pressure of pure component.
 K = Equilibrium constant for a particular component.
 $\Lambda = L/KV$ = absorption factor for a particular component.
 $S = KV/L$ = stripping factor for a particular component.

HEAT QUANTITIES

- H = "Total heat" Enthalpy (above convenient datum temperature) of one mole of vapor.
 h = "Total heat," Enthalpy of one mole of liquid.
 Q_c = Heat quantity removed by condenser in unit time.
 Q_r = Heat quantity supplied by kettle in unit time.
 U = Change in quantity of flowing liquid caused by the introduction of the feed at temperature of feed plate. This is equivalent to the ratio: difference in enthalpy between vapor feed at temperature of feed plate, and feed as delivered to column, divided by molal heat of vaporization at temperature of feed plate

$$= \frac{F(H_f - H_F) + F_L(H_F - h_F)}{H_f - h_f}$$

- T = Absolute temperature °F. + 460.
 t = temperature in degrees Fahrenheit.

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See Mr. Parrish's book on ammonia stills for entrainment.

